

A PROCESS FOR THE REMOVAL OF SO<sub>2</sub>, HCN AND H<sub>2</sub>S AND  
OPTIONALLY COS, CS<sub>2</sub> AND NH<sub>3</sub> FROM A GAS STREAM

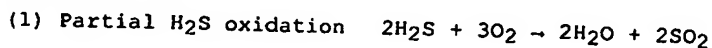
The invention relates to a process for removing SO<sub>2</sub>,  
HCN and H<sub>2</sub>S and optionally COS, CS<sub>2</sub> and NH<sub>3</sub> from a gas  
stream.

Gas streams from industrial chemical processes may  
contain a variety of sulphur compounds such as H<sub>2</sub>S, SO<sub>2</sub>,  
CS<sub>2</sub> and COS. The removal of these sulphur compounds is of  
considerable importance, because the toxicity and smell  
of sulphur compounds renders their presence highly  
undesirable. In addition, other compounds whose presence  
is undesirable may be present such as HCN or NH<sub>3</sub>. It is  
often necessary for the intended purposes of the gas  
stream to remove these compounds at least partly, for  
example to prevent catalyst poisoning in processes after  
the gasification stage.

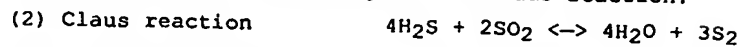
Processes for the removal of sulphur compounds are  
known in the art.

In EP 324,526 a process is described for the removal  
of H<sub>2</sub>S, COS, CS<sub>2</sub> and SO<sub>2</sub> from a gas stream, by converting  
COS and CS<sub>2</sub> into H<sub>2</sub>S. Although EP 324,526 mentions the  
removal of H<sub>2</sub>S, it does not specify the preferred type of  
process for H<sub>2</sub>S removal, nor is the concentration of H<sub>2</sub>S  
in the treated gas stream mentioned.

A process known in the art for the removal of H<sub>2</sub>S  
from a gas stream is the partial oxidation of H<sub>2</sub>S to SO<sub>2</sub>  
according to:



The SO<sub>2</sub> formed can be (catalytically) converted to elemental sulphur according to the Claus reaction:



5 This process is known as the Claus process. The Claus process is frequently employed both in refineries and for the processing of H<sub>2</sub>S recovered from natural gas.

10 A disadvantage of the Claus process is that it is thermodynamically limited by the increase of the water vapour content and simultaneously by the decrease of the H<sub>2</sub>S and SO<sub>2</sub> concentration as the H<sub>2</sub>S conversion proceeds, with the result that the equilibrium of the Claus reaction shifts to the left. Since the dew point of water lies below the solidification point of sulphur, condensation of the water vapour in the process gas to  
15 remove this limitation can lead to serious problems, such as clogging due to the solidification of sulphur and corrosion due to the formation of sulfuric acid, causing serious operational problems.

20 At relatively low concentrations of H<sub>2</sub>S, typically concentrations below 20 vol%, the Claus process is difficult to operate and requires special technical provisions.

25 Another disadvantage of the Claus process is that the concentration of H<sub>2</sub>S in the treated gas is still relatively high. The Claus process has been improved to an extent where the H<sub>2</sub>S content of the treated gas can be lowered to approximately 1 vol% (Superclaus-99 process) and the SO<sub>2</sub> content to approximately 0.02 vol%. However, especially in the case where it is necessary to comply  
30 with requirements with regard to residential use or environmental regulations with respect to emission of sulphur compounds, even lower concentrations of H<sub>2</sub>S and

SO<sub>2</sub>, in the ppm range, have to be achieved. A disadvantage of the Claus and Superclaus process is that such low concentrations of H<sub>2</sub>S and SO<sub>2</sub> cannot be achieved.

5       The removal of HCN from gas streams is important not only because of its own toxic properties, but also in view of corrosive NO<sub>x</sub> compounds which can evolve when both HCN and oxygen are present in a gas stream. In addition, HCN itself is corrosive to equipment when  
10       allowed to concentrate in liquid streams upon gas treating.

      We have now found a relatively simple and inexpensive process for removing SO<sub>2</sub>, H<sub>2</sub>S and HCN and optionally COS, CS<sub>2</sub> and NH<sub>3</sub> from a gas stream. In the process according  
15       to the invention, a combination of reaction steps is used to achieve the removal of sulphur compounds such as H<sub>2</sub>S and SO<sub>2</sub>, and optionally also COS and/or CS<sub>2</sub>, as well as HCN from a gas stream comprising these compounds, to concentrations of below 10 ppmv for H<sub>2</sub>S, below 100 ppmv  
20       for SO<sub>2</sub> and below 0.01 vol%, preferably 10 ppmv for HCN.

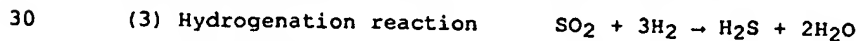
      The invention relates to a process for the removal of SO<sub>2</sub>, HCN and H<sub>2</sub>S and optionally one or more compounds from the group of COS, CS<sub>2</sub> and NH<sub>3</sub> from a first gas stream, which process comprises the steps of:

- 25       (a) removing SO<sub>2</sub> from the first gas stream by contacting the first gas stream in a hydrogenation zone with a hydrogenation catalyst in the presence of hydrogen to obtain a second gas stream;
- 30       (b) removing HCN and optionally COS and/or CS<sub>2</sub> from the second gas stream obtained in step (a) by contacting the second gas stream in a hydrolysis zone with a hydrolysis

catalyst in the presence of water to obtain a third gas stream;

- (c) removing  $\text{NH}_3$  from the third gas stream by contacting the third gas stream in a  $\text{NH}_3$ -removal zone with an aqueous (acidic) washing liquid to obtain an ammonium-comprising aqueous stream and a fourth gas stream;
- (d) removing  $\text{H}_2\text{S}$  from the fourth gas stream by contacting the fourth gas stream in a  $\text{H}_2\text{S}$ -removal zone with an aqueous alkaline washing liquid to obtain a  $\text{H}_2\text{S}$ -depleted gas stream and a hydrogen-sulphide-comprising aqueous stream;
- (e) contacting the hydrogen-sulphide-comprising aqueous stream obtained in step (d) with sulphide-oxidizing bacteria in the presence of oxygen in a oxidation reactor to obtain a sulphur slurry and a regenerated aqueous alkaline washing liquid;
- (f) separating at least part of the sulphur slurry obtained in step (e) from the regenerated aqueous alkaline washing liquid and;
- (g) recycling regenerated aqueous alkaline washing liquid obtained in step (e) to the  $\text{H}_2\text{S}$ -removal zone in step (d).

In step (a) of the process according to the invention,  $\text{SO}_2$  is removed from the first gas stream by contacting the first gas stream in a hydrogenation zone with a hydrogenation catalyst in the presence of hydrogen to obtain a second gas stream. In the hydrogenation zone,  $\text{SO}_2$  is converted to  $\text{H}_2\text{S}$  according to reaction (3):



Reference herein to a first gas stream is to a gas stream comprising  $\text{H}_2\text{S}$ ,  $\text{SO}_2$ ,  $\text{HCN}$  and optionally one or

more compounds selected from the group of COS, CS<sub>2</sub> and NH<sub>3</sub>. The process is especially suitable for first gas streams having a total concentration of sulphur compounds, especially a total concentration of H<sub>2</sub>S, of  
5 between 10 ppmv and 30 vol%, preferably between 100 ppmv and 20 vol%, based on the total gas stream. At these relatively low concentrations of sulphur, especially of H<sub>2</sub>S, conventional processes such as the Claus process cannot be operated without additional technical  
10 provisions, for example amine treatment, to concentrate the H<sub>2</sub>S content, whereas the process according to the invention can be operated without the need for additional technical provisions. Typically, the amount of H<sub>2</sub>S in the first gas stream is at least 50% of the sulphur  
15 compounds, preferably 80%, more preferably 90%.

Usually, the total concentration of HCN in the first gas stream is typically between 10 and 5000 ppmv, suitably between 20 and 1000 ppmv, or between 40 and 250 ppmv, based on the total gas stream.

20 Optionally, solid compounds present in the first gas stream can be removed from the first gas stream prior to entering the hydrogenation zone, for example by use of a filter. This results in less contamination of the sulphur product in step (e) of the process.

25 The hydrogenation zone can be a gas/solid contactor, preferably a fixed bed reactor. Suitable hydrogenation catalysts are for example cobalt/molybdenum, cobalt/tungsten or nickel/molybdenum catalysts. The hydrogenation catalysts suitable for the process  
30 according to the invention can be supported, preferably on a refractory oxide support, more preferably on an alumina support.

Suitably, the temperature in the hydrogenation zone is between 200 °C and 380 °C, preferably between 210 °C and 360 °C, more preferably between 250 °C and 350 °C, still more preferably between 280 and 330 °C. It is found that the hydrogenation reaction proceeds at favourable rates at these temperature ranges. At a temperature range of between 280 and 330 °C, the highest reduction of SO<sub>2</sub> is achieved. The pressure in the hydrogenation zone is suitably between 1 and 100 bara, preferably between 2 and 80 bara.

The hydrogenation reaction results in the second gas stream having a concentration of SO<sub>2</sub> below 100 ppmv, preferably between 0.001 ppmv and 100 ppmv, more preferably between 0.01 ppmv and 50 ppmv, suitably between 0.1 ppmv and 15 ppmv, especially between 0.5 ppmv and 10 ppmv, based on the total gas stream.

In step (b) of the process according to the invention, HCN and optionally COS and/or CS<sub>2</sub> are removed from the second gas stream obtained in step (a) by contacting the second gas stream in a hydrolysis zone with a hydrolysis catalyst to obtain a third gas stream.

In the hydrolysis zone, HCN and, if applicable, COS and/or CS<sub>2</sub> are converted according to the following reactions:

- (4) Hydrolysis of HCN:  $\text{HCN} + \text{H}_2\text{O} \rightarrow \text{NH}_3 + \text{CO}$   
(5) Hydrolysis of COS:  $\text{COS} + \text{H}_2\text{O} \rightarrow \text{H}_2\text{S} + \text{CO}_2$   
(6) Hydrolysis of CS<sub>2</sub>:  $\text{CS}_2 + 2\text{H}_2\text{O} \rightarrow 2\text{H}_2\text{S} + \text{CO}_2$

The amount of water/steam in the hydrolysis zone is preferably between 10 v/v% and 80 v/v%, more preferably between 20 v/v% and 70 v/v%, still more preferably between 30 v/v% and 50 v/v%, based on steam. At the preferred water/steam amounts, the conversion of HCN and

optionally COS and/or CS<sub>2</sub> is better. Typically, the amount of H<sub>2</sub>O in the gas stream is sufficient to achieve conversion of HCN optionally COS and/or CS<sub>2</sub>. Optionally, water or steam of a mixture thereof may be added to the second gas stream prior to contacting the second gas stream with the hydrogenation catalyst, in order to achieve the desired water/steam amount. Optionally, the reaction conditions are selected in such a way, that the reaction mixture remains below the dewpoint of H<sub>2</sub>O. The H<sub>2</sub>O in the gas stream can then advantageously be used for the conversion of HCN and optionally COS and/or CS<sub>2</sub>, to the desired levels.

If COS and/or CS<sub>2</sub> are present, the total concentration of COS and CS<sub>2</sub> in the second gas stream is suitably between 10 ppmv and 2 vol%, preferably between 20 ppmv and 1 vol%, based on the total gas stream.

The hydrolysis zone can be a gas/solid contactor, preferably a fixed bed reactor.

Catalysts for the hydrolysis of HCN and optionally COS and/or CS<sub>2</sub> are known to those skilled in the art and include for example TiO<sub>2</sub>-based catalysts or catalysts based on alumina and/or chromium-oxide. Preferred catalysts are TiO<sub>2</sub>-based catalysts.

Optionally, the hydrolysis step can be done prior to the hydrogenation step. A disadvantage of performing the hydrogenation step prior to the hydrolysis step is that when a TiO<sub>2</sub> catalyst is used, the TiO<sub>2</sub> catalyst can also catalyze the Claus reaction, leading to the formation of sulphur from H<sub>2</sub>S and SO<sub>2</sub>. However, in the case where the temperature is above the dew point of sulphur, the formation of sulphur from the Claus reaction does not lead to clogging.

The hydrolysis results in a third gas stream having a concentration of HCN below 0.01 vol%, suitably between 0.1 ppmv and 0.01 vol%, preferably between 1 ppmv and 50 ppmv, based on the total gas stream.

5       The concentration of COS, if present, in the third gas stream is below 0.01 vol%, suitably between 10 ppmv and 0.01 vol%, preferably between 15 ppmv and 100 ppmv, based on the total gas stream.

10       The concentration of CS<sub>2</sub>, if present, in the third gas stream is below 0.01 vol%, suitably between 1 ppmv and 0.01 vol%, preferably between 2 ppmv and 50 ppmv, based on the total gas stream.

15       In step (c) of the process according to the invention NH<sub>3</sub> is removed from the third gas stream by contacting the third gas stream in an NH<sub>3</sub>-removal zone with an aqueous (acidic) washing liquid to obtain an ammonium-comprising aqueous stream and a fourth gas stream. The process is especially suitable for a third gas stream having an amount of NH<sub>3</sub> of between 10 and 20       6000 ppmv, preferably between 20 and 2000 ppmv. The temperature in the NH<sub>3</sub>-removal zone is suitably between 5 and 70 °C, preferably between 10 and 50 °C, to achieve a sufficient removal of NH<sub>3</sub> at a low temperature. The pressure in the NH<sub>3</sub>-removal zone is suitably between 1 25       and 100 bara, preferably between 2 and 80 bara, to achieve a sufficient removal of NH<sub>3</sub> at a low pressure.

30       In step (d) of the process according to the invention H<sub>2</sub>S is removed from the fourth gas stream by contacting the fourth gas stream in H<sub>2</sub>S-removal zone with an aqueous alkaline washing liquid to obtain a H<sub>2</sub>S-depleted gas stream and a hydrogen-sulphide-comprising aqueous stream.

The process is especially suitable if the load of



sulphur compounds in the H<sub>2</sub>S-removal zone is below  
 60000 kg/day, suitably between 50 and 50000 kg/day,  
 preferably between 75 and 20000 kg/day, more preferably  
 between 100 and 10000 kg/day. At these sulphur loads,  
 5 conventional processes such as the Claus process are  
 difficult, if not impossible, to operate, whereas the  
 process according to the invention can be used  
 advantageously.

Suitably, the total amount of H<sub>2</sub>S of the fourth gas  
 10 stream is between 10 ppmv and 20 vol%, preferably between  
 20 ppmv and 10 vol%. An advantage of the process  
 according to the invention is that the H<sub>2</sub>S in the fourth  
 gas stream can be removed even when the H<sub>2</sub>S amount is  
 relatively low, typically between 10 ppmv and 20 vol%.  
 15 For other processes such as the Claus process it is  
 necessary that a sour gas is produced that has a high H<sub>2</sub>S  
 content to make it suitable as a Claus feed, for example  
 by an amine treatment and regeneration step prior to the  
 Claus reaction.

20 Suitable aqueous washing liquids include aqueous  
 hydroxide solutions, e.g. sodium hydroxide or potassium  
 hydroxide solutions in water. The pH of the aqueous  
 alkaline solvents is suitably between 7 and 12,  
 preferably between 8 and 11.

25 The main reactions that can take place in the  
 H<sub>2</sub>S-removal zone in step (d) are:

- |                                 |  |
|---------------------------------|--|
| (7) H <sub>2</sub> S absorption | $\text{H}_2\text{S} + \text{OH}^- \rightarrow \text{HS}^- + \text{H}_2\text{O}$  |
| (8) H <sub>2</sub> S absorption | $\text{H}_2\text{S} + \text{CO}_3^{2-} \rightarrow \text{HS}^- + \text{HCO}_3^-$ |
| (9) CO <sub>2</sub> absorption  | $\text{CO}_2 + \text{OH}^- \rightarrow \text{HCO}_3^-$                           |
| 30 (10) Carbonate formation     | $\text{HCO}_3^- + \text{OH}^- \rightarrow \text{CO}_3^{2-} + \text{H}_2\text{O}$ |
| (11) Poly-hydrosulphide         | $2\text{HS}^- + \text{S}_8 \rightarrow 2\text{HS}_5^-$                           |

formation

The term sulphide-comprising aqueous stream as used herein refers to an aqueous stream comprising one or more products of the main reactions (7) to (11) that can take place in the H<sub>2</sub>S-removal zone, such as HS<sup>-</sup>, disulphides, polysulphides, thiocarbonates and carbonates but can also include dissolved H<sub>2</sub>S.

The preferred temperature in the H<sub>2</sub>S removal zone is between 5 and 70 °C, more preferably between 10 and 50 °C. Preferably, the pressure in the H<sub>2</sub>S removal zone is between 1 and 100 bara, more preferably between 2 and 80 bara.

Typically, the H<sub>2</sub>S removal zone is a gas/liquid contactor. Suitable gas/liquid contactors are described in Perry's Chemical Engineers' Handbook, 7th edition, section 14 (1997) and include for example a tray or packed column or a gas scrubber.

Optionally, the medium of the H<sub>2</sub>S removal zone is buffered. Preferred buffering compounds are carbonates, bicarbonates phosphates and mixtures thereof, especially sodium carbonate and/or sodium bicarbonate. The concentration of the buffering compounds depends inter alia on the composition of the gas flow and is generally adjusted in such a way, that the pH of the reaction medium in the H<sub>2</sub>S removal zone is between 6.0 and 10, more preferably between 6.5 and 9.0.

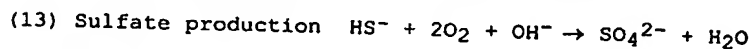
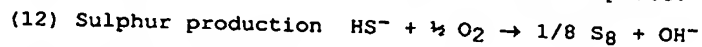
Optionally, fresh alkaline washing liquid can be supplied to the H<sub>2</sub>S removal zone to achieve the desired pH.

Due to their odorous nature, H<sub>2</sub>S, mercaptans, sulphides, disulphides and aromatic mercaptans can be detected at parts per million concentrations. Thus, it is

desirable for users of such gas and refinery streams to have total concentration of sulphur compounds, especially H<sub>2</sub>S, lowered to a concentration of e.g. less than 30 or 20 ppmv, preferably less than 10 ppmv, based on the total H<sub>2</sub>S-depleted gas stream.

The process results in a "H<sub>2</sub>S-depleted gas stream" having a total concentration of sulphur compounds, especially H<sub>2</sub>S, of below 35 ppmv, suitably between 0.01 and 30 ppmv, or below 25 ppmv, suitably between 0.01 and 20 ppmv, or below 15 ppmv, suitably between 0.01 and 10 ppmv, preferably between 0.05 and 3.5 ppmv, more preferably between 0.1 and 1 ppmv, based on the total gas stream.

In step (e) of the process according to the invention the hydrogen-sulphide-comprising aqueous stream obtained in step (d) is contacted with sulphide-oxidizing bacteria in the presence of oxygen in a oxidation reactor to obtain a sulphur slurry and a regenerated aqueous alkaline washing liquid. The main reaction that can take place in the oxidation reactor in step (e) is the microbiological formation of sulphur and sulphate:



The term sulphur slurry as used herein refers to a slurry comprising one or more products of the main reactions, including reactions (12) and (13), that can take place in the oxidation reactor.

The term regenerated aqueous alkaline solvent as used herein refers to an aqueous alkaline solvent wherein at most 2 w/w% sulphur particles are present, preferably at most 1 w/w%.

The amount of oxygen fed into the oxidation reactor is adjusted such that the oxidation of absorbed sulphide results predominantly in sulphur, as suggested in NL 8801009, disclosing a process for the controlled oxidation of sulphur-containing waste water.

Reference herein to sulphide-oxidizing bacteria is to bacteria which can oxidize sulphide to elemental sulphur. Suitable sulphide-oxidizing bacteria can be selected for instance from the known auto tropic aerobic cultures of the genera Thiobacillus and Thiomicrospira.

Typical pressures in the oxidation reactor in step (e) are between 1 and 2 bara.

Suitably, the oxidation reactor has a volume of between 5 and 2500 m<sup>3</sup>, preferably between 10 and 2000 m<sup>3</sup>.

Preferably, the reaction medium in the oxidation reactor in step (e) is buffered. The buffering compounds are chosen in such a way that the bacteria present in the oxidation reactor tolerate them. Preferred buffering compounds are carbonates, bicarbonates phosphates and mixtures thereof, especially sodium carbonate and/or sodium bicarbonate. The concentration of the buffering compounds depends inter alia on the composition of the gas flow and is generally adjusted in such a way, that the pH of the reaction medium in the oxidation reactor is between 6 and 10, more preferably between 7 and 9.

In step (f) of the process according to the invention at least part, typically between 5 and 95 w/w%, preferably between 10 and 90 w/w% based on the total weight of the slurry, of the sulphur slurry obtained in step (e) is separated from the regenerated aqueous alkaline solvent. The phrase "at least part of" as used herein also includes a complete separation of sulphur from the regenerated aqueous alkaline liquid.

Suitably, the separating step takes place in a solid/liquid separator. Suitable solid/liquid separators are described in Perry's Chemical Engineers' Handbook, 7th edition, section 22 (1997).

5 Typically, the sulphur content of the separated aqueous sulphur slurry is between 5 w/w% and 50 w/w%, based on the slurry. Typically, the water of the sulphur slurry is removed to an extent that a sulphur cake with a dry solids content of between 55 and 70% is obtained.  
10 Typically, the sulphur content of the sulphur cake is between 90 and 98 w/w%, based on the total weight of the sulphur cake. Optionally, the sulphur slurry obtained in step (c) can be re-slurried, filtered and dried to obtain a sulphur paste with a purity of at least 95 wt% sulphur,  
15 preferably at least 99 wt% sulphur. The sulphur paste thus-obtained can optionally be dried to produce a powder with a dry weight content of at least 85%, preferably at least 90%. This powder can suitably be applied as a fungicide or as a miticide.

20 The sulphur produced in the process according to the invention has a hydrophilic nature and does not cause the fouling problems that are typically caused by sulphur produced by non-biological liquid processes. Another advantage of the sulphur produced in the process  
25 according to the invention is that it is very suitable for use as a fertilizer.

In step (g) of the process according to the invention, regenerated aqueous alkaline washing liquid obtained in step (e) is recycled to the H<sub>2</sub>S-removal zone  
30 in step (d). Suitably, between 10 and 99%, preferably between 30 and 95%, more preferably between 40 and 90% of the total amount of regenerated aqueous alkaline washing liquid obtained in step (e) is recycled to the

H<sub>2</sub>S-removal zone in step (d). By recycling the regenerated aqueous washing liquid to the H<sub>2</sub>S-removal zone fresh aqueous washing liquid is supplied to the H<sub>2</sub>S-removal zone for the removal of H<sub>2</sub>S. This enhances the removal of H<sub>2</sub>S to a concentration of 30 ppmv or less, suitably 20 ppmv or less, preferably 10 ppmv or less. The regenerated aqueous alkaline washing liquid optionally comprises sulphur particles.

The process according to the invention is especially suitable for the treatment of a gaseous CO-containing stream, especially a syngas stream, an offgas stream from a carbon black plant, or a refinery gas stream. Syngas is a general term that is applied to mixtures of carbon monoxide, hydrogen, inert components and, carbon dioxide that are derived from the gasification of coal, oil residues, waste or biomass. The main components of syngas are hydrogen and carbon monoxide. Further, often carbon dioxide and traces of methane are present. Refinery streams concern crude oil derived gaseous streams containing smaller or larger amounts of sulphur compounds. Recycle streams and bleed streams of hydrotreatment processes, especially hydrodesulfurisation processes, can also suitably be treated by the process according to the invention.

The invention will now be illustrated by means of schematic figure 1. Figure 1 depicts a typical process scheme according to the invention, wherein a gas stream comprising SO<sub>2</sub>, HCN, H<sub>2</sub>S and optionally COS, CS<sub>2</sub> and NH<sub>3</sub> enters via line (1) the hydrogenation zone (2) via line (1) and is contacted with the hydrogenation catalyst to remove the SO<sub>2</sub>. Hydrogen is supplied to the hydrogenation zone via line (3). After treatment in the hydrogenation zone the second gas stream, which is

SO<sub>2</sub>-depleted, leaves the hydrogenation zone via line (4) and enters the hydrolysis zone (5) where it is contacted with a hydrolysis catalyst to remove HCN and optionally COS and/or CS<sub>2</sub>. Optionally, steam or a steam/water mixture can be added to the hydrolysis zone via line (6). After treatment in the hydrolysis zone, the third gas stream, which is depleted of HCN, COS and/or CS<sub>2</sub>, leaves the hydrolysis zone via line (7) and enters the HN<sub>3</sub>-removal zone (8), where it is treated with aqueous (acidic) washing liquid to obtain an ammonium-comprising aqueous stream and a fourth gas stream. Optionally, steam or a steam/water mixture can be added to the HN<sub>3</sub>-removal zone via line (9). After treatment in the NH<sub>3</sub>-removal zone, the fourth gas stream, which is depleted of HCN, COS and/or CS<sub>2</sub> and NH<sub>3</sub>, leaves the HN<sub>3</sub>-removal zone via line (10) and enters the H<sub>2</sub>S-removal zone (11), where it is contacted with an aqueous washing liquid. The H<sub>2</sub>S-depleted gas stream is led from the H<sub>2</sub>S-removal zone via line (12), optionally via a knockout vessel (13) and can be further processed via line (14). A sulphide-comprising aqueous stream leaves the H<sub>2</sub>S-removal zone via line (15) and enters a flash vessel (16). H<sub>2</sub>S-depleted excess gas is vented off from the flash vessel through line (17). In the case that flash gas contains some H<sub>2</sub>S, the H<sub>2</sub>S is removed in a small flash gas contactor before further use as fuel gas. The sulphide-comprising aqueous stream is led via line (18) to the aerobic reactor (19) where the sulphide compounds are oxidized. Nutrients and air are fed to the aerobic reactor via feed lines (20) and (21). The sulphur-containing solid/liquid mixture generated in the aerobic reactor is led via line (22) to a solid/liquid separator (23) where the sulphur-slurry is

separated and discharged via line (24). The liquid is led back to the bioreactor via line (25). Off-gas is vented from the aerobic reactor via line (26). The regenerated aqueous alkaline solvent is led to the H<sub>2</sub>S-removal zone via line (27).

The invention will now be illustrated by the following non-limiting examples.

EXAMPLE 1 (comparative)

A gas stream containing sulphur- compounds and HCN with concentrations as given in table 1 is heated to a temperature of 280 °C and led to a reactor filled with a stacked bed of hydrogenation catalyst and hydrolysis catalyst. The gas is cooled in a quench tower at 40 °C and treated in an amine/Claus unit to remove the H<sub>2</sub>S. The concentrations of sulphur compounds and of HCN in the treated gas is given in table 1.

Table 1

	Gas to treating system (v/v%)	Gas after treating system (v/v%)
H <sub>2</sub> O	6.7	3.2
H <sub>2</sub> S	1.9	0.01
SO <sub>2</sub>	0.1	0.05
CS <sub>2</sub>	0.09	0.01
COS	0.28	0.01
HCN	0.03	0.01
NH <sub>3</sub>	0.04	0.03

EXAMPLE 2 (according to the invention)

A first gas stream containing sulphur- compounds, NH<sub>3</sub> and HCN with concentrations as given in table 2 is heated to a temperature of 280 °C and led to a reactor filled



with a stacked bed of hydrogenation catalyst and hydrolysis catalyst to create a second gas stream and subsequently a third gas stream. The third gas stream is cooled in a quench tower at 40 °C and treated in an NH<sub>3</sub>-scrubber to create a fourth gas stream. The fourth gas stream is treated in a H<sub>2</sub>S-removal unit to create a H<sub>2</sub>S-depleted gas stream. The resulting hydrogensulphides are oxidised in a bioreactor. The concentrations of the sulphur-containing compounds and of the HCN in the treated gas (H<sub>2</sub>S-depleted gas) are given in table 2.

Table 2

	Gas to treating system (first gas stream) (v/v%)	Gas after treating system (H <sub>2</sub> S-depleted gas stream) (v/v%)
H <sub>2</sub> O	40	6.8
H <sub>2</sub> S	0.9	0.0002
SO <sub>2</sub>	0.1	0.001
CS <sub>2</sub>	0.04	0.0002
COS	0.01	0.001
HCN	0.1	0.001
NH <sub>3</sub>	0.03	0.001

From the examples it is clear that with the process according to the invention a substantially lower concentration of sulphur-containing compounds, especially of H<sub>2</sub>S, SO<sub>2</sub>, CS<sub>2</sub> and COS is achieved, while the concentration of NH<sub>3</sub> and HCN is also lowered.